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Techno-economic prospects for CO_2 capture from distributed energy systems

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ABSTRACT

CO_2 emissions from distributed energy systems are expected to become increasingly significant, accounting for about 20% for current global energy-related CO_2 emissions in 2030. This article reviews, assesses and compares the techno-economic performance of CO_2 capture from distributed energy systems taking into account differences in timeframe, fuel type and energy plant type. The analysis includes the energy plant, CO_2 capture and compression, and distributed transport between the capture site and a trunk pipeline. Key parameters, e.g., capacity factor, energy prices and interest rate, were normalized for the performance comparison.

The findings of this study indicate that in the short-mid term (around 2020–2025), the energy penalty for CO_2 capture ranges between 23% and 30% for coal-fired plants and 10–28% for natural gas-fired plants. Costs are between 30 and 140 €/tCO_2 avoided for plant scales larger than 100 MW_{LHV} (fuel input) and 50–150 €/tCO_2 avoided for 10–100 MW_{LHV}. In the long-term (2030 and beyond), the energy penalty for CO_2 capture might reduce to between 4% and 9% and the costs to around 10–90 €/tCO_2 avoided for plant scales larger than 100 MW_{LHV}, 25–100 €/tCO_2 avoided for 10–100 MW_{LHV} and 35–150 €/tCO_2 avoided for 10 MW_{LHV} or smaller.

CO_2 compression and distributed transport costs are significant. For a distance of 30 km, 10 €/tCO_2 transported was calculated for scales below 500 tCO_2/day and more than 50 €/tCO_2 transported for scales below 5 tCO_2/day (equivalent to 1 MW_{LHV} natural gas). CO_2 compression is responsible for the largest share of these costs.

CO_2 capture from distributed energy systems is not prohibitively expensive and has a significant cost reduction potential in the long term. Distributed CO_2 emission sources should also be considered for CCS, adding to the economies of scale of CO_2 transport and storage, and optimizing the deployment of CCS.

1. Introduction

CO_2 capture and storage (CCS) is considered to become an important technology to mitigate anthropogenic global warming. It is currently expected that CCS will be deployed mainly in the power sector and the industry, in which many large scale point sources are found [44]. Smaller scale energy systems, however, may account for a considerable fraction of total energy-related CO_2 emissions in the future. Distributed generation (DG) is expected to become increasingly important in the future energy supply infrastructure, particularly in future electric utilities in the economies where deregulation takes place [84]. DG stations are generally smaller than 100–150 MW_e [1] and combined heat and power generation (CHP) is one of the major applications of DG due to its high overall energy efficiency. A study made by the International Energy Agency (IEA) estimates that in a scenario to halve the global GHG emissions in 2050, compared to 2007 level (BLUE Map scenario), 30% of global electricity generation from fossil fuels would come from gas-fired CHP plants [46]. Moreover, The BLUE Map scenario also estimates that more than a quarter of gas-fired CHP electricity generation, or 8% of global electrical generation from fossil fuels, will be equipped with CO_2 capture in 2050.

Distributed energy systems are generally not considered for CCS due to the relatively high costs at such scales [22,59]. However, there are a number of differences between DG plants and centralized power plants that may challenge this claim. First, the operating conditions of DG plants are different from those for centralized power plants that the economies of scale may not necessarily apply. For example, CHP plants may be able to achieve a high degree of heat integration with CO_2 capture

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process that centralized power plant cannot, such as the recovery of low-temperature heat. Moreover, the literature indicates that CO_2 capture from distributed energy systems would become economical in the longer term. A study by the IEA Greenhouse Gas R&D Programme (IEA GHG), which performed a screening of technologies for "medium scale" (1–100 MW_{th}) distributed energy systems and an order-of-magnitude analysis for the economic performance of five selected technologies [55], identifies solid oxide fuels (SOFC) and oxyfuel coal boilers using oxygen conducting membranes (OCM) to be potentially promising. Damen et al. [21] shows that CO_2 capture from a 20 MW_e SOFC system may enable CO_2 avoidance cost as low as those for large-scale systems (600 MW_e) in the long term. Therefore, it is important to quantify the performance development potential for CO_2 capture in distributed energy systems.

Second, no study to date has assessed and compared the performance of CO_2 capture in distributed energy systems taking into account the diversity in type of generator technologies used, applications, plant scales, operational patterns and applicable CO_2 capture technologies. There are a considerable number of CO_2 capture technologies proposed in the literature. A broad assessment is therefore desired to obtain better insights into the techno-economic possibilities for distributed CO_2 capture.

Third, there are few studies that provide insights into the minimum scale of CO_2 emission sources to which CO_2 capture can be applied economically. Several studies on CCS use a 100 $ktCO_2$/yr as the scale limit e.g., [20,54,59,105]. The IPCC Special Report on CCS excludes emissions from sources smaller than 100 $ktCO_2$/yr with the argument that their emissions represent only a small fraction of the total CO_2 emissions [59]. The 100 $ktCO_2$/yr limit, therefore, seems to be a practical limitation and not a technical or economic limitation.

In this context, the objectives of this study are two-fold: (1) to provide an overview of techno-economic performance of CO_2 capture from distributed energy systems in the short-mid term (ST/MT) and the long-term future (LT), which explicitly accounts for differences in timeframe, fuel type and energy plant type, and (2) to assess the relative cost of CO_2 capture from distributed generation compared to large-scale centralized power plants, taking into account the differences in plant scale, operational conditions, and the combination of CO_2 compression and distributed transport. This study focuses on CO_2 capture from fossil fuel-fed distributed energy systems. While there are other potentially promising distributed CO_2 capture options such as CO_2 capture from atmosphere using a solid sorbent for sustainable hydrocarbon fuel production [36], they are not considered in this study.

This paper is structured as follows. Section 2 provides an overview of distributed CO_2 emission sources and the type of technologies used for decentralized energy conversion systems. This is followed by a description of the key methodological aspects used in this study (Section 3). Sections 4 and 5 assess the techno-economic performance CO_2 capture technologies for decentralized emission sources for the ST/MT and the LT, respectively. For the ST/MT, the number of commercially feasible CO_2 capture technologies is limited. We investigated how the energy penalty for a particular ST/MT CO_2 capture technology would differ by the type and application of energy systems. In the LT, in contrast, there is a variety of advanced technologies and concepts proposed in the literature. We therefore put emphasis on the assessment of the technology itself. Section 6 combines the results from Sections 4 and 5 to assess the technological and economic performance improvement potential from ST/MT to LT. Section 7 assesses the effect of CO_2 capture scale on compression and distributed transport costs. Section 8 discusses the limitations of the study while conclusions are drawn in Section 9.

2. CO_2 emissions from distributed energy systems

In this research, three main types of distributed energy systems are considered: CHP plants, boilers and distributed hydrogen plants. This section briefly describes the technology, applications, and global CO_2 emissions of each type. Note that while this study will include the three technologies as far as the availability of data allows it, the focus will be on CHP plants.

2.1. CHP plants

The main advantage of CHP is the high overall energy efficiency due to the simultaneous production of useful thermal and electrical energy [26]. The type of installed energy systems depends on the heat demand and the quality of heat product required. Table 1 presents an overview of heat demands by sector and the applicability of various combined heat and power (CHP) generator types.

While CHP enables large energy savings and CO_2 emissions reduction, its CO_2 emissions are still significant. CHP accounts for more than 6 EJ_e/yr, or more than 10% of global electricity generation today [42], and is expected to increase considerably in the coming decades. An IEA study suggests that in G8+5 countries,[1] which account for more than two-thirds of global primary energy consumption, the share of CHP in electricity generation may increase from 11% in 2005 to 24% in 2030 in a scenario with a pro-CHP policy regime [45]. Assuming the same share worldwide, CHP would account for about 23 EJ_e/yr[2] and over 3 $GtCO_2$ emissions per year using conventional technology and conservative assumptions.[3]

2.2. Boilers

The IEA GHG [55] estimates that boilers with a fuel input of 1–100 MW_{th}, [4] which corresponds to the typical scale of district heating (DH), industrial and large commercial installations, account for about 9% of global energy-related emissions (more than 2 $GtCO_2$/yr). The majority of emissions is attributable to coal boilers in China [55].

Assuming that global CO_2 emissions from boilers remain constant, distributed energy systems (CHP plants and boilers combined) in 2030 could account for about 20% of current global energy-related CO_2 emissions[5] .

2.3. Distributed hydrogen plants

Hydrogen may gradually take over the function of natural gas in the residential sector in the longer term [22]. Barreto et al. [6]

[1] G8+5 countries include G8 nations (Canada, France, Germany, Italy, Japan, Russia, the United Kingdom and the United States) plus Brazil, China, India, Mexico and South Africa.

[2] The CHP potentials reported in the IEA [45] are based on the global electricity generation figures of the Alternative Policy Scenario (APS) in the World Energy Outlook 2007 [43]. APS assumes 12% reduction in global electricity generation in 2030 compared to the reference scenario, which estimates 29,737 TW h/yr. Therefore, global electricity generation in 2030 under APS is 94 EJ_e/yr. Hence, $94 \times 24\% = 22.6$ EJ_e/yr.

[3] All CHP plants around the world were assumed to be natural gas-fired with an electrical efficiency of 40%.

[4] A 100 MW_{th} coal-fired plant operating for 7500 h/yr would produce about 250,000 t/yr of CO_2 and a natural gas-fired plant would produce about 150,000 t/yr, assuming an emission factor of 95 kg/GJ_{LHV} and 56 kg/GJ_{LHV}, respectively [53].

[5] Note that the increased penetration of CHP would affect global boiler capacity. For example, CO_2 emissions from Chinese coal boilers may reduce drastically in the future because the Chinese government is closing down coal-fired power plants with capacities smaller than 100 MW_e to reduce coal consumption and CO_2 emissions [112].

Table 1
Overview of combined heat and power (CHP) plant technologies.
Source: EPA [26], IEA [45], IEA GHG [55] and IPCC [59].

CHP system	Typical capacity [MW_e]	Power efficiency (LHV) (%)[a]	Overall efficiency (LHV) (%)[a]	Typical heat to power ratio	Installed cost [$€_{2007}/kW_e$]	O&M costs [$€_{2007}/kWh_e$]	Fuel types	Uses for thermal output	Exhaust gas CO_2 concentration [v/v, dry]
Gas turbine	0.5–40 (NGCC: ~250)	24–40	77–83	0.5–2	710–950 (5–40 MW_e)	0.003–0.01	NG, biogas, propane, oil	Heat, hot water, LP–HP steam	3–4% (NG)
Reciprocating engine	0.01–5	24–44	77–88	1–2	800–1600	0.007–0.02	NG, biogas, propane, landfill gas	Hot water, LP steam	3–4% (NG) [b]
Steam turbine	0.5–250	16–42	84–88	3–10	310–800	< 0.004	All types	LP–HP steam	12–14% (coal)
Fuel cells	0.005–2	33–66	72–88	0.5–1	370–4800	0.02–0.03	H_2, NG, propane, methanol	Hot water, LP–HP steam	Technology dependent [c]

[a] Efficiency values reported in high heating value (HHV) are converted to low heating value (LHV) terms using a multiplication factor of 1.05 for coal and 1.1 for other fuels.
[b] IEA GHG [55] presents a higher range (9–14%), but we consider the range an overestimation. Gas engine exhaust gas typically contains 11–16% O_2 [61], leading to 3–5 vol% (dry) CO_2.
[c] A node off-gas of high temperature fuel cells (e.g., solid oxide fuel cells (SOFC) and molten carbonate fuel cells (MCFC)) is CO_2-rich when fuels rich in hydrocarbons or carbon monoxide are used. However, the anode off-gas contains some unoxidized fuel that it cannot be compressed and transported for underground storage. In conventional SOFC and MCFC systems, anode off-gases are fully oxidized with air. Therefore, CO_2 concentrations of plant exhaust gases are low; only 2–3 vol% (dry) for SOFC [8,69] and around 7 vol% (dry) for MCFC [10]. IEA GHG [55] presents somewhat higher values (8–10%), but the report does not provide references.

present an future energy scenario in which CHP in industrial and residential stationary fuel cell applications and electricity generation from mobile hydrogen-based fuel cells in the transportation sector together account for nearly 40% of the global electricity generation in 2100. Some fuel cell technologies may become economical for CHP on a house/building (block) level for newly build projects [22].

In addition, a significant development of hydrogen markets might be seen if society takes a path for hydrogen-powered vehicles, e.g., fuel cell vehicles. In this case, large-scale hydrogen plants using fossil feedstock are considered the most economical, but the cost of building a distribution infrastructure and the need for sufficient demand could be major barriers. Therefore, distributed hydrogen generation at refueling stations using the existing natural gas infrastructure may be considered [95].

3. Research methods

3.1. Data collection, system boundaries and reference system selection

We reviewed literature that investigated the techno-economic performance of CO_2 capture from energy systems of scales up to about 150 MW_{th} fuel input for electricity and heat supply systems and up to 2–3 MW_{th} output for hydrogen plants. This scale limit covers "medium scale" DG systems, which is defined to be of 5–50 MW_e scale by Ackermann et al. [1]. For hydrogen plants, the aforementioned plant scale is representative of future hydrogen refueling stations for hydrogen vehicles [95].[6] Regarding the economic performance data available in the literature, we only considered those from studies published after year 2000 to take into account the latest developments in CO_2 capture R&D and to avoid economic data which are outdated. In this study, in principle only newly built plants were considered. Nevertheless, performance data for retrofit cases reported in the literature were also used when they prove to be useful for this study.

The system boundaries defined in this study included the following components: energy conversion plant, CO_2 capture and compression, and distributed CO_2 transport between the CO_2 capture site and trunk CO_2 pipeline. All plants were assumed to be newly-built and trunk CO_2 pipelines were assumed to be already available. The CO_2 capture technologies assessed in this study were categorized into ST/MT technologies (10–15 years) and LT (20 years or more) technologies. ST/MT technologies are defined as those that are either in pilot, demonstration or commercialization phase today [86]. The technologies are categorized as ST/MT also when all required components are commercially available today, even if the process as a whole has not been tested or demonstrated. All other technologies, either in modeling or laboratory phase today, are considered to be LT options.

The selection of the reference system has a large impact on the assessment of the CO_2 capture performance. Throughout this study, we applied the common approach comparing identical energy conversion plants with and without CO_2 capture [21]. This approach enables to obtain a clear insight into which technologies may enable inherent cost-effective CO_2 capture [21].

[6] Note that for 50 MW H_2 scale plants, Meerman et al. [78] investigated cost-effective CO_2 capture from natural gas-fed steam methane reformers using existing technologies.

3.2. Technical performance indicator

We used the efficiency penalty of an energy plant type *i* (condensing power plant, CHP, H_2, boiler, furnace) due to CO_2 capture ($\Delta\eta_{pen}$: dimensionless) as an indicator of technical performance of CO_2 capture, which is expressed as:

$$\Delta\eta_{pen} = 1 - \frac{\eta_{CC}}{\eta_{Ref}} \quad (1)$$

where η is the net energy conversion efficiency of the plant. Subscripts *CC* stands for the plant with CO_2 capture and Ref for the reference plant, respectively. For CHP plants, η is calculated on exergy terms:

$$\eta = \frac{E_{el,out} + E_{th,out} \times f}{E_{PE}} \quad (2)$$

where $E_{el,out}$ is the net electrical output (MW_e), $E_{th,out}$ is the net heat product output (MW_{th}), *f* is the exergy factor for heat, and E_{PE} is the primary energy input (MW_{LHV}).

The exergy content method provides a thermodynamically acceptable allocation base between steam and electricity by acknowledging the higher usefulness of electricity [99]. The exergy-based allocation is also suggested to be the most meaningful and accurate among various methods for allocating CO_2 emissions for cogeneration processes [91].

In this study, the energy performance of DH CHP plants was calculated using the data for condensing plants, taking into account various opportunities for heat recovery. Fig. 1 shows a simplified diagram of a district heating CHP plant with CO_2 capture and compression, and possible heat recovery opportunities. The energy performance of DH CHP plants with and without CO_2 capture were compared on leveled heat and power outputs. $E_{el,out}$ and $E_{th,out}$ were calculated as follows:

$$E_{el,\,out} = E_{el,cond} - E_{el,comp} - E_{el,CC} - E_{th,CC} \times f_{CC} - E_{th,DH} \times f_{DH} \quad (3)$$

$$E_{th,out} = E_{th,DH} + E_{th,CC} \times r_{th} + \left(E_{el,CC} + E_{el,comp}\right) \times r_{el} \quad (4)$$

where $E_{el,cond}$ is the net power output at condensing mode (power-only mode) (MW_e), $E_{el,comp}$ is the power consumption for CO_2 compression (MW_e), $E_{el,CC}$ is the power consumption for CO_2 capture (MW_e), $E_{th,CC}$ is the heat extracted for CO_2 capture (MW_{th}), f_{CC} is the exergy factor for CO_2 capture heat (dimensionless), $E_{th,DH}$ is the heat extracted for DH (MW_{th}), r_{th} is the fraction of CO_2 capture heat recovered for DH, and r_{el} is the fraction of CO_2 capture and compression power consumption recovered as heat for DH.

For boilers, η is calculated as:

$$\eta = \frac{E_{th,\,out}}{E_{PE} + \left(E_{el,\,in}/\eta_{grid}\right)} \quad (5)$$

where $E_{el,in}$ is the net electricity consumption (MW_e), and η_{grid} is the centralized electricity generation efficiency. For hydrogen plants, η is calculated based on primary energy terms as proposed in Damen et al. [21]:

$$\eta = \frac{E_{H_2} + \left(E_{el,\,out}/\eta_{grid}\right) + \left(E_{th,\,out}/\eta_{boiler}\right)}{E_{PE} + \left(E_{el,\,in}/\eta_{grid}\right) + \left(E_{th,\,in}/\eta_{boiler}\right)} \quad (6)$$

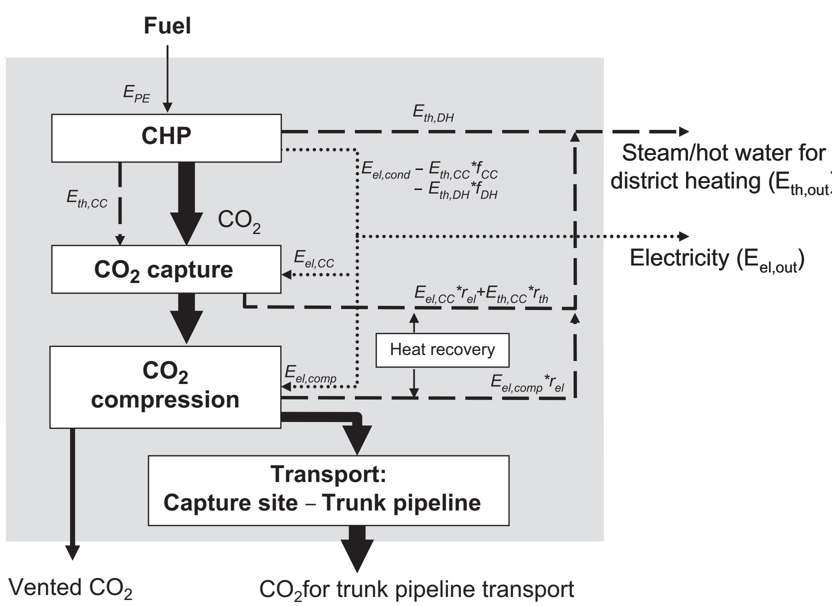


Fig. 1. District heating CHP plant with CO_2 capture and compression. The figure also shows possible heat recovery opportunities.

Table 2
Exergy factors for various heat products and centralized electricity generation efficiency.

Parameter	Symbol	Value
Exergy factor for heat (f: J_{ex}/J_{th}) [a]		
District heating (steam or hot water)	f_{DH}	0.17
CO_2 capture solvent regeneration heat (low temperature steam)	f_{CC}	0.22
Industrial process heat (medium–high temperature steam)	f_{Ind}	0.35
Centralized electricity generation efficiency	η_{grid}	0.5

[a] DH, CO_2 capture solvent regeneration heat and industrial steam were assumed to be extracted at 110 °C, 140 °C, and 240 °C, respectively. The exergy factors were taken from a figure in Bolland and Undrum [7].

where $E_{th,in}$ is the net steam consumption (MW_{th}), η_{boiler} is the boiler efficiency (90%).

Table 2 presents exergy factors for various heat qualities, including centralized electricity generation efficiency (η_{grid}). The exergy factors for heat product, *f*, are differentiated by the temperature of the heat product.

3.3. Economic performance indicator

In this study, we used CO_2 avoidance cost (C_{CO_2}: €/tonne) as the main economic performance indicator. CO_2 avoidance cost was calculated based on the production cost of an energy carrier (*COE*: €/GJ):

$$C_{CO_2} = \frac{COE_{CC} - COE_{Ref}}{Em_{Ref} - Em_{CC}} \quad (7)$$

where *Em* is the specific CO_2 emission intensity of an energy carrier (tCO_2/GJ). For CHP plants, the *COE* largely depends on how costs are allocated to the electricity and heat that are produced. In this study, we used the exergy factor of final energy carriers to allocate costs. *COEs* are calculated by dividing the sum of annualized capital cost, O&M costs and fuel costs by the annual production of the energy carrier.

3.4. Normalization of parameters

All cost figures were converted to €$_{2007}$. Inflation and material price increases for the energy systems and CO_2 capture process plants were accounted for by applying the CERA European Power Capital Cost Index (EPCCI) [57]. Costs that are reported in US$ were first converted to Euro of the original year using the exchange rate data (year average) from OANDA [83], then updated to €$_{2007}$. The Chemical Engineering Plant Cost Index CEPCI; [16] was used for distributed CO_2 transport components. Other cost figures were corrected for inflation by using GDP deflators [58]. Note that for advanced CO_2 capture technologies, technical performance figures are forecasts and cost projections are uncertain [21].

With regard to capital investment, we considered total capital requirement (TCR), which includes the following components:

- Process plant cost (costs for the equipment pieces and their installation) plus engineering fees and contingencies;
- Owner costs (royalties, preproduction costs, inventory capital, land costs and site preparation) and interests during construction.

3.4.1. Key parameters

Table 3 presents the parameters that are normalized for technical and economic performance calculations throughout the study. High and low parameter values were used to assess the sensitivity of the results (see Section 3.5).

3.4.2. Normalization of CO_2 compression, distributed transport and CO_2 purity

Literature indicates that the use of trucks maybe more suitable than pipelines for small-scale and short-distance CO_2 transport [40,109]. We therefore considered both branch pipeline and truck as options for distributed CO_2 transport between the CO_2 capture site and a trunk pipeline (see Fig. 2). In each case, the cheapest

Table 3
Parameters normalized for technical and economic performance calculations in this study.

Parameters and variables	Unit	Nominal	Low	High
Real interest rate	%	10	7.5	12.5
Economic lifetime of plants[a]				
Coal-fired installations	yr	35	30	40
Gas-fired installations (conventional technologies)	yr	25	20	30
Membrane-based technologies	yr	20	15	25
Annual operation time				
Industrial installations	h/yr	7500	6000	8500
Centralized power plants	h/yr	7000	6000	8000
Other installations[b]	h/yr	5000	3500	6500
Total capital requirement [c]	%—total plant cost	110	105	130
Total plant cost[c]	%—process plant cost	130	120	150
Energy prices				
Coal[d]	€/GJ_{LHV}	2.6	2	3.2
Natural gas[d]	€/GJ_{LHV}	8	5	11
Electricity [e]	€/MW h	55	40	70
CO_2 emission factors				
Natural gas [f]	g/MJ_{LHV}	56	–	–
Coal [f]	g/MJ_{LHV}	95	–	–
Grid electricity ($Em_{Sp,Elec}$)[g]	g/MJ_e	111	89	133

[a] Lifetime for membrane-based technologies, e.g., fuel cells and membrane reactors, are based on [30,75].
[b] "Other installations" include district heating, institutional and commercial installations.
[c] Process plant cost (PPC) comprises equipment cost and installation costs. Total plant cost (TPC) comprises PPC and engineering fees and contingencies. Total capital requirement (TCR) comprises TPC, owner costs and interests during construction. The values are within the ranges observed for power plant construction [106].
[d] Nominal value is from IEA [47]. The high and low values assumed here agree with those forecasted by the IEA for the EU, the US and Japan for years between 2020 and 2030 [47].
[e] Electricity price for large industries differs significantly by country, from 0.028 €/kWh in Russia (in 2006) to 0.177 €/kWh in Italy (in 2007). The price used in this study is similar to that in the USA (0.048 €/kWh in 2007), South Korea (0.052 €/kWh in 2007) and Poland (0.061 €/kWh in 2007) [103].
[f] From [60].
[g] The changes in electrical consumption due to CO_2 capture are likely to affect base load power generation by base-load fossil fuel-fired power plants. We assumed that in the industrialized world where CCS is also deployed for industrial processes, around 40% of base-load fossil fuel-fired power plants are equipped with CO_2 capture. Assuming an average 75% CO_2 avoidance rate by CO_2 capture, we estimated that the CO_2 emissions from base-load fossil fuel-fired power plants would be reduced by 30% due to CO_2 capture. The nominal value assumes that natural gas- and coal-fired power plants share the electricity generation 50% each. The low end value corresponds to a ratio of 80:20 between natural gas power plants and coal power plants, and the high end value corresponds to a ratio of 20:80.

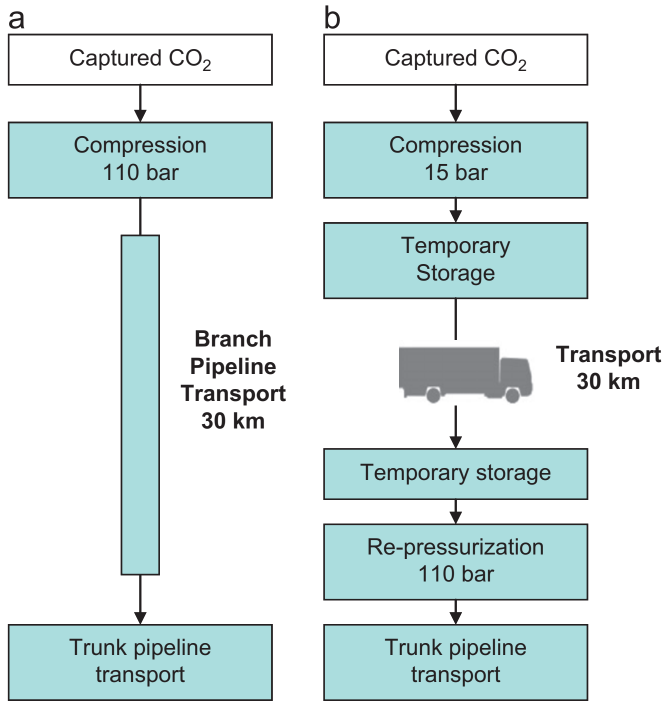


Fig. 2. Distributed CO_2 transport options considered in this study.

option was selected. A distributed transport distance of 30 km was assumed and we did not consider the networking of clustered distributed energy systems for collective CO_2 capture. Note that these assumptions are conservative for industrial areas or residential areas in which CO_2 capture-distributed energy systems are expected to be installed.[7] For truck transport, CO_2 is transported in trucks in liquid phase, typically at 17 bar, −30 °C [109].[8] The CO_2 delivered by truck is re-pressurized for trunk pipeline transport. In this study, we assumed that total CO_2 compression cost (onsite compression and re-compression at the trunk pipeline inlet) for truck transport is equivalent to that for onsite CO_2 recompression to 110 bar. This assumption can be justified because CO_2 compression is performed in multiple stages (often more than four) and there would be limited effect of scale on capital costs and electricity consumption.

For both branch transport options, CO_2 is compressed to 110 bar for pipeline transport in a supercritical phase. CO_2 compression is performed in two steps [77]. First, CO_2 is compressed using a multistage compressor up to the critical pressure of CO_2 (7.38 MPa), followed by a further compression using a pump for liquid/supercritical phase. When electricity consumption for CO_2 compression is not reported in the original literature, specific power consumption is estimated using the following equation adapted from Koornneef et al. [66]:

$$E_{s,\mathrm{comp}} = \frac{ZRT_1}{M\eta_{is}\eta_m}\frac{N\gamma}{\gamma-1}\left\{\left(\frac{p_c}{p_{in}}\right)^{\gamma-1/N\gamma}-1\right\}+\frac{p_{out}-p_c}{\rho\eta_P} \quad (8)$$

where $E_{s,\mathrm{comp}}$ is the specific electricity requirement (kJ/kg CO_2), Z is the average CO_2 compressibility factor[9] (0.89), R is the universal gas constant (8.3145 J/(mol K)), T_1 is the suction temperature (303.15 K), γ is the specific heat ratio (c_p/c_v) (1.294), M is the molar mass (44.01 g/mol), η_{is} is the isentropic efficiency (80%), η_m is the mechanical efficiency (99%), p_{in} is the suction pressure (101 kPa), p_c is the critical pressure (7380 kPa), p_{out} is the outlet pressure (110 bar = 11,000 kPa), N is the number of compressor stages (= 4), ρ is the density of CO_2 during pumping (630 kg/m^3), and η_P is the pump efficiency (75%). When the compression pressure reported in the literature differs from the value used in this study (110 bar), we also adjusted the specific electricity consumption using Eq. (8). The pressure loss in branch pipelines was assumed negligible in this study.

CO_2 compressor cost was calculated using the parameters presented in Table 4. The CO_2 compression system capital cost (installation cost) was calculated using a regression curve fitted to data from the following sources [67,71,104,110]:

$$C_{\mathrm{comp}} = 2.7 \times E_{\mathrm{el,comp}}^{0.53} \quad \left(n=4,\quad R^2=0.98,\quad 0.11 < E_{\mathrm{el,comp}} < 13\right) \quad (9)$$

where C_{comp} is the capital cost for the CO_2 compressor (M €), $E_{el,\mathrm{comp}}$ is the power consumption for CO_2 compression (MW$_e$), n: is the number of CO_2 compression system capital cost data, and R^2 is the coefficient of determination.

CO_2 compression may have a significant impact on the economic performance of smaller scale CO_2 capture [55]. We therefore looked carefully into the CO_2 compressor costs reported in the literature reviewed in this study. If those assumed in the literature were considered too low, we recalculated the compressor capital costs using Eq. (9).

In case the CO_2 purity reported in the literature is lower than 95 vol%,[10] an additional CO_2 purification process was assumed to achieve 95 vol% purity. In such cases, CO_2 capture rate was adjusted by a multiplication factor η_{Rec} because some CO_2 would be vented together with the removed impurities in the purification process. η_{Rec} was assumed 90%, 92% and 94% for CO_2 purities below 75 vol%, between 75 and 80 vol%, and above 80 vol%, respectively [90].

For branch pipeline transport cost calculations, the model presented in IEA GHG [50] was used (see also [68][11]). The parameters used for truck transport cost calculations are presented in Table 4. Besides the costs for truck operation itself, the costs for CO_2 storage tanks at the capture site for truck transport also needs to be taken into account.

3.5. Sensitivity analysis

A sensitivity analysis was conducted for the following parameters: energy prices, plant lifetime, interest rate, capital cost, and grid electricity CO_2 emission factor (see Table 3). To assess the combined effect of the parameters on the results, we applied a general equation for uncertainty propagation as described in,

[7] Large industrial or urban areas in Europe where CCS is considered, e.g., Randstad (includes Amsterdam and Rotterdam) in the Netherlands and the Merseyside and Deeside Basin (includes Liverpool) in the UK, fit within a radius of 30–40 km [22,56].

[8] This means that captured CO_2 is not compressed to 110 bar as assumed in the previous section. Nevertheless, studies indicate that specific power consumption and capital cost are comparable for liquefaction at low pressures (15–20 bar) and at critical pressure. Studies on ship transport at 50 °C and 6.5–7 bar report specific electricity consumption of 110–123 kW h/tCO_2 [5,52]. The reported capital cost is 80M\$$_{2004}$ for 102 MW$_e$ power, which is in agreement with the capital cost equation for the systems for 110 bar (Eq. (9)). Therefore, we assumed that the onsite electricity consumption calculated for branch pipeline transport is also valid for truck transport.

[9] The derived value is the average of the compressibility factor values calculated for the average pressure of each compressor stage (four stages in total). Compressibility factors for each compression stage were calculated using the Peng–Robinson equation of state programmed by ChemSOF [17].

[10] 95 vol% is a typical concentration at which existing CO_2 pipelines operate [23].

[11] Except for a terrain factor which was reduced from 2 to 1.4 based on the argument in van den Broek et al. [105].

Table 4
Parameters for CO_2 compression and distributed CO_2 transport cost calculations.

Parameter	Unit	Value
Truck transport cost [a]	€$_{2007}$/(t km)	0.22
CO_2 re-pressurization system capital cost[b]	M€$_{2007}$	1.2 × compression power (MW_e)+0.07
CO_2 storage tank cost (200 tCO_2 capacity)[c]	k€$_{2007}$	150
O&M cost: pipeline and CO_2 storage tank[b]	%—capital cost	2

Because of the assumed transport distance, CO_2 boil-off during transport and CO_2 emissions from transport trucks were assumed to be negligible. The CO_2 boil-off rate can be as high as 10% depending on the length of time the CO_2 is kept in the truck [109]. For ocean tankers, a boil-off rate of 1% per day is suggested [39].

[a] From van Bergen et al. [104]. The values reported in the literature range between 0.04 €/(t km) [79] and 0.32 €/(t km) [64]. Although it is not explicit in the literature, the low estimate seems to take into account only the truck operation cost (driver and fuel), and the high estimate seems to include CO_2 liquefaction. Values that lie within the range are reported in Wong [109], Herzog and Golomb [40] and MGSC [79].

[b] From McCollum and Ogden [77]. The minimum storage capacity was assumed to be 20 t, which is equivalent to a typical capacity of CO_2 transport trucks.

[c] From Wong [109]. The scaling factor was assumed to be 0.6. The minimum tank capacity was assumed to be 20 t CO_2, equivalent to the truck transport capacity.

e.g., [82,101]. In this approach, for a given indicator C, which is a function of variables $X_1, X_2,\ldots,X_n$, the standard deviation of C (σ_C) is calculated as follows:

$$\sigma_C = \sqrt{\sum_{i=1}^{n}\left(\frac{\partial C}{\partial Xi}\right)^2 \sigma_{Xi}{}^2 + 2\sum_{i=1}^{n-1}\sum_{j=i+1}^{n}\frac{\partial C}{\partial Xi}\frac{\partial C}{\partial Xj}\sigma_{Xi}\sigma_{Xj}r_{ij}} \quad (10)$$

where σ_{Xi} is the standard deviation of variable X_i, r_{ij} is the correlation coefficient between two variables X_i and X_j. Because there is no detailed information on the distribution of the parameter values, we assumed a uniform distribution. In this case, σ_{Xi} is calculated as [82]:

$$\sigma_{Xi} = \frac{X_{i,\mathrm{High}} - X_{i,\mathrm{Low}}}{\sqrt{12}} \quad (11)$$

The correlation coefficients among variables were not considered in this study.

4. Assessment of short-mid term CO_2 capture technologies

As noted earlier, we investigated how the energy penalty for a particular ST/MT CO_2 capture technology would differ by the type and application of energy systems. Then we looked at a number of promising CO_2 capture options for distributed energy systems that have not yet been covered in earlier studies.

4.1. Literature review and selection of short-term technologies

Table 5 shows the key performance parameters for ST/MT CO_2 capture from small–medium scale emission sources reported in the literature. Most of the data are for industrial applications.

4.1.1. Post-combustion capture

Post-combustion capture in the ST/MT would most likely apply chemical absorption process. Chemical absorption is more favorable for power plants with steam turbines than for boilers and simple gas turbine power plants because steam turbines can minimize the exergy losses caused by the extraction of steam for CO_2 capture. General description of chemical absorption process can be found in, e.g., [59]. Flue gas is first cooled down from about 110 °C to about 50 °C before it enters the CO_2 scrubber. CO_2 is then bound by the chemical solvent in the scrubber at temperatures between 40 and 60 °C. After the CO_2 absorption process, the CO_2-rich solvent flows through the stripper at temperatures around 100–140 °C, where the chemical bounded CO_2 and is removed and the solvent is regenerated. CO_2 is released at the top of the stripper with considerable amount of steam. In condensing power plants, the steam is condensed and separated from CO_2 in a water cooler. The hot CO_2-lean solvent is pumped back to the stripper via lean-rich heat exchanger and a cooler to bring the temperature down to 40–60 °C.

In case of capture from natural gas combustion flue gas, MEA-based process consumes around 20–30% of total thermal energy input for solvent regeneration. For boilers, this directly leads to $\Delta\eta_{\mathrm{pen}}$ of 20–30% (Table 5). For NGCC plants, however, $\Delta\eta_{\mathrm{pen}}$ is much smaller (e.g., IEA GHG [51] reports 7.5%) because the MEA process requires low-temperature steam, which has low exergy content.

4.1.1.1. District heating CHP plants. To the authors' knowledge, the study by Desideri and Corbelli [24] on a 5 MW_e steam-injection gas turbine plant is one of the few in the public literature that assesses the techno-economic feasibility of retrofitting CO_2 capture to smaller scale DH CHP plants. The results show that the energy penalty differs significantly by the operational mode of the CHP plant [24]. There are a number of studies that investigate the integration between large scale DH CHP plants and CO_2 capture e.g., [48,65]. These studies suggest that chemical absorption CO_2 capture may be energetically more economical for DH CHP plants than for condensing power plants because of better heat integration possibilities.

In the chemical absorption CO_2 capture process, low-grade heat can be recovered from mainly three sections: flue gas cooling, CO_2 regenerator condenser, and lean solvent cooling. Regarding flue gas cooling, heat is not recovered in condensing power plants because the temperature is too low for the low-pressure evaporator [72], but it is sufficiently high to generate hot water [24]. For the CO_2 regenerator condenser, the IEA GHG [48] has shown that in case of NGCC plants, about 30% of the thermal energy extracted from the steam turbines for CO_2 capture can be recovered at the CO_2 regenerator condenser in the form of hot water[12] when the DH water returns at 40 °C. Knuutila et al. [65] and Gode and Hagberg [34] report similar recovery rates.[13] It is suggested that this heat recovery may be done with negative

[12] 55 MW_{th} can be recovered for a plant with 854 MW_{th} fuel input and 90% CO_2 capture. Assuming a CO_2 emission factor of 56 g/MJ, specific heat recovery is calculated to be 1.28 GJ/tCO_2 captured. The IEA GHG study assumes specific heat consumption of 4.22 GJ/tCO_2 captured, thus 1.28/4.22=0.303.

[13] The figure from the Knuutila study also includes heat recovery from CO_2 compression, but we considered this fraction in total heat recovered to be small. 107 MW of hot water can be generated at 108 kgCO_2/s capture rate and a specific heat consumption of 3.22 GJ/tCO_2 captured. The heat recovery rate is calculated to be 30.7%. Bode and Hagberg [34] report that the electrical efficiency reduces by 11.4%-points while DH efficiency increases by 8.9%-points. Using the exergy factor presented in Table 2 (f_{CC}=0.22) and assuming that CO_2 compression accounts for

Table 5
Energy, economic and CO_2 performance data for small–medium scale installations with short-mid term CO_2 capture technologies reported in the literature.

DG category	Fuel type	Generator type	Plant scale [MW output]	CO_2 capture technology	CO_2 capture efficiency (%)	System energy efficiency (η_{Sys}) (%)	Energy penalty ($\Delta\eta$) (%)	Capital cost [$€_{2007}$/kW output]	O&M cost [%—Cap. cost]	Reference
Condensing/CHP plants	NG	GT	5	Chemical absorption	68–90	26–32	19–25	3800[a]	7.3	[24]
			5	Pre-combustion	85	41	19	2000	2.6	[55]
		NGCC (industrial, partial load)	50[a]	Chemical absorption	90	46–53	12–18[b]	1500–1600[a]	4	[68]
Boilers	NG	Boiler	140	Chemical absorption	86	70	28	400	2.9	[100]
	Coal	Boiler	50	Chemical absorption	85	58	33	990	3.6	[55]
	Refinery gas	Boiler	78–110	Oxyfuel (stand-alone ASU)	90	22–25	30–40	1200–1580	6.8[c]	[113]
Furnaces	Refinery gas	Various refinery furnaces	17–103	Oxyfuel (stand-alone ASU)	90	68–77	6–23	620–1150 (incremental cost) [c]	1.8–3.1	[108]

[a] Plant scale is expressed in rated electrical capacity. Capital cost is expressed in $€_{2007}$/kW rated electrical capacity in maximum power generation mode.
[b] The values are presented for cases 1A and 2A. Both cases are operated at 60% of the maximum fuel input, and case 1A is operated at 25% heat efficiency and 2A at 40% efficiency. The high energy penalty ($\Delta\eta$) value was calculated for case 2A and the low $\Delta\eta$ value was calculated for case 1A.
[c] Cost data for air-blown furnaces without CO_2 capture are not available.

incremental capital cost because the direct contact cooling tower is replaced by additional heat recovery steam generator surface, the cost of which is marginal [48]. For lean solvent cooling, a considerable amount of heat is released in the lean amine cooling process, but heat recovery from the additional cooling is unrealistic because the temperature difference between the cooler inlet and outlet is small ([48] assumes 27 °C).

4.1.1.2. Industrial CHP plants. To date, few feasibility studies have been published on post-combustion CO_2 capture from small–medium scale industrial CHP plants, in particular taking into account the operational conditions. For NGCC–CHP plants, our previous study [68],[14] shows that the energy penalty for plants in partial load operation (about 60%) may become 13–15% lower than that for condensing plants when the CHP plant is operating at a low heat efficiency of 25%. This is mainly due to a better use of the capacity of the NGCC–CHP plants. By increasing the fuel input rate to meet energy requirements for CO_2 capture, the plant efficiency increases and partly offsets the CO_2 capture energy requirements. In such cases, in the ST/MT CO_2 avoidance costs for NGCC–CHP plants at scales as small as 100 MW_e (condensing plant scale) may become lower than those for 400 MW_e condensing plants. However, energy penalty becomes 40% higher than that for power-only NGCC plants when the CHP plant is operating at a higher heat efficiency of 40%, because supplementary firing will be required to meet heat demand for CO_2 capture.

Chemical absorption CO_2 capture from simple gas turbine (GT) CHP plants is considered to be uneconomical. Most GT–CHP plants are smaller than 40 MW_e (Table 5), which is indicated to be expensive for CO_2 capture from NGCC–CHP plants even when making the best of unused capacity [68]. Moreover, heat extraction from GT–CHP plants results in larger exergy losses compared to that from NGCC–CHP plants. Therefore, this option is not investigated further in this study.

4.1.2. Oxyfuel combustion CO_2 capture

Studies on the application of oxyfuel combustion CO_2 capture to smaller scale emission sources are scarce. There are some pilot oxyfuel coal boiler plants up to 30 MW_{th} scale (e.g., Schwarze Pumpe in Germany) being tested to date [44]. Reduction of the CO_2 capture energy penalty can be achieved through heat integration in DH CHP plants. Low-grade heat can be recovered from the ASU and CO_2 compression unit. A case study on a German coal fluidized bed power plant [94] shows that more than 20% of power consumption for CO_2 capture can be recovered in the form of DH and capital cost can be reduced by 5% compared to condensing plants.

With regard to gas-fired power plants, oxyfuel combustion capture for natural gas-fired power plants is reported to be considerably more expensive than other CO_2 capture technologies (e.g., [53,62]) because special gas turbines have to be used for gas-fired oxyfuel power plants [49]. Therefore, ST/MT oxyfuel combustion CO_2 capture for natural gas-fired installations is not investigated further in this study. Relatively cheap CO_2 capture, however, may become possible for carbon-intensive gases such as refinery gases of certain composition. Moreover, CO_2 emission sources are clustered in a refinery such that collective CO_2 may

(*footnote continued*)
3.6%-point electrical efficiency reduction (based on specific electricity consumption of 0.4 GJ_e/tCO_2), the heat recovery rate is 8.9/{(11.4–3.6)/0.22}=25.1%.

[14] This study also made estimates for mid-term future (2020–2025) based on the future technological development estimated by Peeters et al. [86]. However, we in recent years consider the estimates to be on the optimistic side taking into account the developments in CCS R&D in recent years. Therefore, the mid-term future results are not considered here.

Table 6
Potential energy and capital cost reduction as a result of heat integration for ST/MT CO_2 capture technologies in district heating and industrial applications.

Fuel	Application	Post-combustion (MEA)	Oxyfuel
NG	DH	~30% solvent regeneration heat recovery	No information available. Possibly heat recovery from ASU and CO_2 compression, and capital cost reduction due to system integration.
	Industrial	NGCC at partial load operation: ~15% energy penalty reduction	
Coal	DH	~30% solvent regeneration heat recovery	~25% of ASU and CO_2 compression power recovered as DH ~5% capital cost reduction for CHP
	Industrial	No particular benefits	No particular benefits

enable cost reduction due to economies of scale [13,27]. In particular, ASU can benefit significantly from upscaling.

4.1.3. Pre-combustion capture

For coal-fired plants, integrated gasification combined cycle (IGCC) is a state-of-the-art power generation technology that could potentially enable low-cost CO_2 capture in the longer term [63]. However, pre-combustion CO_2 capture would not be the most economical option in the ST/MT, especially for distributed energy systems for mainly two reasons. First, pre-combustion capture does not use any component that is of modular nature, so it is unlikely that it would have scale advantages over other CO_2 capture technologies at smaller plant scales. Second, IGCC technology is still in the early phase of market introduction and it is unlikely that small–medium scale IGCC power plants would be built for district heating in the ST/MT.

For natural gas-fired plants, pre-combustion capture is reported to be significantly more expensive than other CO_2 capture processes for 500 MW_e scale condensing plants [62]. However, for smaller scale systems such as simple GT–CHP plants, pre-combustion may enable lower energy penalty compared to chemical absorption capture. The IEA GHG [55] proposes a 5 MW_e system using autothermal reforming (ATR), water–gas shift reactor and pressure swing adsorption (PSA) for CO_2 capture.

Table 6 presents the applicability of ST/MT CO_2 capture technologies investigated and assessed in this section. Our literature review indicated that the following smaller scale CO_2 capture options may be economical when fitted to DH CHP plants: (1) chemical absorption for NGCC–CHP, and (2) chemical absorption and (3) oxyfuel combustion for PC–CHP plants. In the following section, the techno-economic performance of DH CHP plants with CO_2 capture is investigated in detail.

4.2. Performance of ST/MT CO_2 capture technologies for DH CHP plants

Chemical absorption and oxyfuel combustion were considered for coal-fired CHP plants and chemical absorption was considered for NGCC–CHP plants. As noted in Section 3.2, full-scale coal-fired power plants with and without CO_2 capture were scaled down to one-tenth calculate the performance for smaller scale DH CHP plants.

A net heat to power ratio (HPR) of 1 was considered for PC–CHP plants and 0.5 for NGCC–CHP plants, respectively. For NGCC–CHP plants, it is unlikely that chemical absorption CO_2 capture will be fitted to plants operating at high HPR because of the competition for heat and the limit to the HPR that NGCC–CHP plants can reach. There are two options to enable relatively high net HPR while achieving high CO_2 capture efficiency (85–90%): (1) use auxiliary boiler for additional heat, or (2) build a larger CHP plant with increased electricity generation capacity, but both of these options may not be economical. The first option is not recommended because of increase in CO_2 capture energy penalty since only low-grade heat is generated from a high quality fuel. A significant cost increase in such configurations has been reported in [68]. The second option requires larger capital investments and the increased electricity generation can become problematic if the electricity market is unfavorable for exporting excess electricity to the grid. In addition, for DG systems there is a technical limit of the grid to accept the electrical capacity [1] that generating large amount of excess electricity may not be favorable.

4.2.1. Technical and economic parameters

The technical and economic performance for DH–CHP plants with ST/MT CO_2 capture was calculated based on the performance data for full-scale condensing plants. To ensure that their performance data are comparable, the base performance data for coal-fired [81] and natural gas-fired plants [80] were taken from a set of reports prepared by the same group (NETL) published between 2007 and 2008. Coal-fired plants were assumed to be bituminous coal-fed supercritical plants.[15]

Regarding the calculation of technical and economic performance for smaller scale plants, Table 7 shows the effect of scale on the technical performance of power plant and CO_2 capture components.

The maximum total gross power and heat efficiency of CHP plants was limited at 95%. In case the CHP plant cannot meet CO_2 capture energy demand for the nominal CO_2 capture efficiency due to the limit in maximum gross power and heat efficiency or the minimum gross power efficiency, CO_2 capture rate was adjusted to the heat that can be generated.

Capital costs for the full-scale power plants, with and without CO_2 capture, were broken down into key components. A scaling factor for each component was applied to calculate total capital costs at smaller plant scales (Eq. (12)):

$$C_X = \sum_i \left\{ C_{\mathrm{Ref},i} \times \left(\frac{S_{X,i}}{S_{X_{\mathrm{Ref}},i}} \right)^{SF_i} \right\} \quad (12)$$

where C_X is the capital cost for plant scale X (MW_e), $C_{\mathrm{Ref},i}$ is the capital cost for the component i for the reference (full-scale) plant, $S_{X,i}$ is the capacity of component i for the plant scale X, $SX_{\mathrm{Ref},i}$ is the capacity of component i for the reference plant scale X_{Ref}, and SF_i is the scaling factor for component i.

Table 8 shows the capital cost scaling factors for various components of power plants and CO_2 capture processes investigated in this study.

4.2.2. Results

Table 9 shows the performance results for coal-fired DH CHP plants with and without CO_2 capture for different plant scales (full-scale and one-tenth scale) for the operation at a net heat-to-power ratio (HPR) of 1. The results indicate slightly better net CHP efficiency for chemical absorption capture than for oxyfuel combustion capture. The effect of increased specific power

[15] Although supercritical plants are not common for small scales today, we expect it will become by the time CCS becomes commercial.

Table 7
Effect of scale on the technical performance of power plant and CO_2 capture components.

Parameter	Assumption and references
Coal fired power plant efficiency	Unaffected by plant scale. This assumption agrees with the plant efficiencies calculated by the IECM model [14].
NGCC efficiency	Varies directly with the 0.0679th power of gross power capacity as reported in Kuramochi et al. [68], which was derived from data provided in the Gas Turbine World Handbook 2007–2008 [37].
Chemical absorption capture specific energy requirement	Unaffected by plant scale. This assumption is consistent with the assumptions made in previous studies for NGCC plants of 6–62 MW_e scale [33] and coal-fired steam turbine plants of 100–2500 MW_e scale [14].
ASU specific power consumption (GJ_e/tO_2)	Varies directly with the -0.0474th power of plant scale (tO_2/d). The value 0.0474 is estimated from a figure that depicts the correlation between specific power consumption and plant scale presented in Castle [12].

Table 8
Capital cost scaling factors.

Equipment	Scaling based on	Scaling factor	References and notes
Flue gas desulphurization	Gross power capacity under condensing operation	0.72	[92]
Coal-fired power plant	Gross power capacity under condensing operation	0.77 [a]	[14]
NGCC	Gross power capacity under condensing operation	Varies directly with $(732+1.1X)/(1+0.0052X)$ (X: gross power capacity in MW_e)	[68]
De-NO_X (selective catalytic reduction)	Flue gas volume	0.7	[61]
Chemical absorption CO_2 capture unit	CO_2 capture capacity	0.6	[88]
Air separation unit	Oxygen separation capacity	0.85[b]	[38]
CO_2 compression and purification	Compression power	0.53	See Eq. (9)

[a] For 100–2000 tO_2/d.
[b] The capital costs were calculated by the IECM model to be 1852 $/$kW_e$ net for 96.4 MW_e net plant and 1275 $/$kW_e$ net for 482 MW_e net plant.

Table 9
Performance results for district heating (DH) coal-fired CHP plants with and without CO_2 capture in the ST for the operation at a net heat-to-power ratio (HPR) of 1. The performance data are also shown for full-scale condensing plants (cond.) from NETL [80,81], on which the calculations were based.

CO_2 capture technology	No capture		Chemical absorption				Oxyfuel			
Plant scale	Full scale		Full scale		One-tenth scale		Full-scale		One-tenth scale	
Net HPR	0 (cond.)	1	0 (cond.)	1	0 (cond.)	1	0 (cond.)	1	0 (cond.)	1
Net electrical output ($E_{el,out}$: MW_e)	550	470	549	470	53	47	549	470	53	47
Gross electrical efficiency	43.1%	37.2%	35.3%	32.5%	35.3%	32.5%	43.4%	39.4%	43.4%	39.6%
System power consumption										
Power plant	−2.2%	−2.2%	−2.5%	−2.2%	−2.5%	−2.2%	−2.1%	−2.1%	−2.1%	−2.1%
Post-combustion CO_2 capture plant	–	–	−1.1%	−1.1%	−1.1%	−1.1%	–	–	–	–
Air separation unit	–	–	–	–	–	–	−7.0%	−7.0%	−7.8%	−7.8%
CO_2 Compression and purification	–	–	−2.4%	−2.4%	−2.4%	−2.4%	−4.0%	−4.0%	−4.0%	−4.0%
Net electrical efficiency	40.8%	34.9%	29.3%	26.8%	29.3%	26.8%	30.4%	26.4%	29.6%	25.7%
DH heat efficiency ($E_{th,out}/E_{PE}$)	–	34.9%	–	26.8%	–	26.8%	–	26.4%	–	25.7%
Electrical efficiency reduction due to CO_2 capture thermal energy extraction [$(E_{th,CC}\times r_{CC})/E_{PE}$]	–	–	7.8%	7.8%	7.8%	7.8%	–	–	–	–
CO_2 capture heat recovered as DH heat ($E_{th,CC}\times r_{th}/E_{PE}$)	–	–	–	10.7%	–	10.7%	–	2.8%	–	3.0%
Gross energy conversion efficiency	43.1%	72.1%	64.6%	84.2%	64.6%	84.2%	43.4%	65.8%	43.4%	65.3%
Net total CHP efficiency [$(E_{el,out}+E_{th,out})/E_{PE}$]	–	69.8%	–	53.6%	–	53.6%	–	52.7%	–	51.4%
Energy penalty ($\Delta\eta_{pen}$)	–	–	28.2%	23.2%	28.2%	23.2%	25.6%	24.5%	27.6%	26.4%
Capital cost (€$_{2007}$/kW_e net)	1270	1486	2296	2562	4531	5122	2161	2373	3941	4350
O&M cost (%—capital cost)	3.1%	2.6%	2.9%	2.3%	1.9%	1.6%	2.6%	2.2%	1.9%	1.6%

consumption for ASU on the net CHP efficiency was found to be limited for the plant scales investigated in this study. In comparison with the plants in condensing operation, the efficiency penalty for chemical absorption CO_2 capture decreased by 18% from 28.2% to 23.2%, while for oxyfuel combustion CO_2 capture the decrease was marginal (4%).

Table 10 shows the performance results for DH NGCC–CHP plants with and without CO_2 capture for different plant scales (full-scale and one-tenth scale) for the operation at a net heat-to-power ratio (HPR) of 0.5. In comparison with the plants in condensing operation, $\Delta\eta_{pen}$ decreased by 25%. $\Delta\eta_{pen}$ became larger for smaller plants (12.2% for one-tenth scale plant compared to 10.5% for full-scale plant) because the energy efficiency reduction in percentage-point terms was the same for full-scale and one-tenth scale (8.1%-points) while the reference system efficiency was lower for one-tenth scale plants.

Fig. 3 shows CO_2 avoidance costs for DH–CHP plants with chemical absorption and oxyfuel CO_2 capture for varied plant scales. Costs are broken down into components to show the effect of factors contributing to the difference in CO_2 avoidance costs between full scale condensing plants and one-tenth scale CHP plants. For coal-fired CHP plants, CO_2 avoidance costs were similar for the two capture technologies at full-scale but oxyfuel capture became significantly cheaper than chemical absorption capture at smaller scales because of lower capital costs. The cost reduction due to heat recovery for post-combustion capture case was found to be 10 €/tCO_2 avoided. For NGCC–CHP plants operating at net HPR of 0.5, the figure shows that the cost reduction due

Table 10
Technical performance results for district heating (DH) NGCC CHP plants with and without CO_2 capture for different plant scales for the operation at a net heat-to-power ratio (HPR) of 0.5. The performance data are also shown for full-scale condensing plants (cond.) from NETL [80,81], on which the calculations were based.

CO_2 capture technology	No capture			Chemical absorption capture		
Plant scale	Full scale		One-tenth scale	Full scale		One-tenth scale
Net HPR	0 (cond.)	0.5	0.5	0 (cond.)	0.5	0.5
Net electrical output ($E_{el,out}$: MW_e)	560	516	52	482	516	52
Gross electrical efficiency	57.4%	53.0%	45.3%	52.4%	50.1%	42.4%
System power consumption: Total	−1.0%	−1.0%	−0.8%	−3.9%	−3.5%	−3.4%
Power plant	−1.0%	−1.0%	−0.8%	−1.4%	−1.0%	−0.9%
Post-combustion CO_2 capture plant	–	–	–	−1.0%	−1.0%	−1.0%
Air separation unit	–	–	–	–	–	–
CO_2 compression and purification	–	–	–	−1.5%	−1.5%	−1.5%
Net electrical efficiency ($E_{el,out}/E_{PE}$)	56.4%	52.0%	44.5%	48.5%	46.6%	39.0%
DH heat efficiency ($E_{th,out}/E_{PE}$)	–	26.0%	22.2%	–	23.3%	19.5%
Electrical efficiency reduction due to CO_2 capture thermal energy extraction [$(E_{th,CC}\times r_{CC})/E_{PE}$]	–	–	–	5.0%	5.0%	5.0%
CO_2 capture heat recovered as DH heat ($E_{th,CC}\times r_{th}/E_{PE}$)	–	–	–	–	6.9%	6.9%
Gross energy conversion efficiency [a]	57.4%	79.0%	67.6%	–	89.4%	78.0%
Net total CHP efficiency [$(E_{el,out}+E_{th,out})/E_{PE}$]	–	78.0%	66.7%	–	69.9%	58.6%
Energy penalty ($\Delta\eta_{pen}$)	–	–	–	14.0%	10.5%	12.2%
Capital cost (€$_{2007}$/kW$_e$ net)	445	483	864	942	1017	1520
O&M cost (%—capital cost)	2.8%	2.4%	2.0%	2.4%	2.4%	1.5%

[a] Gross energy conversion efficiency is defined as the sum of gross electricity output and gross thermal product ouput divided by the total primary energy input.

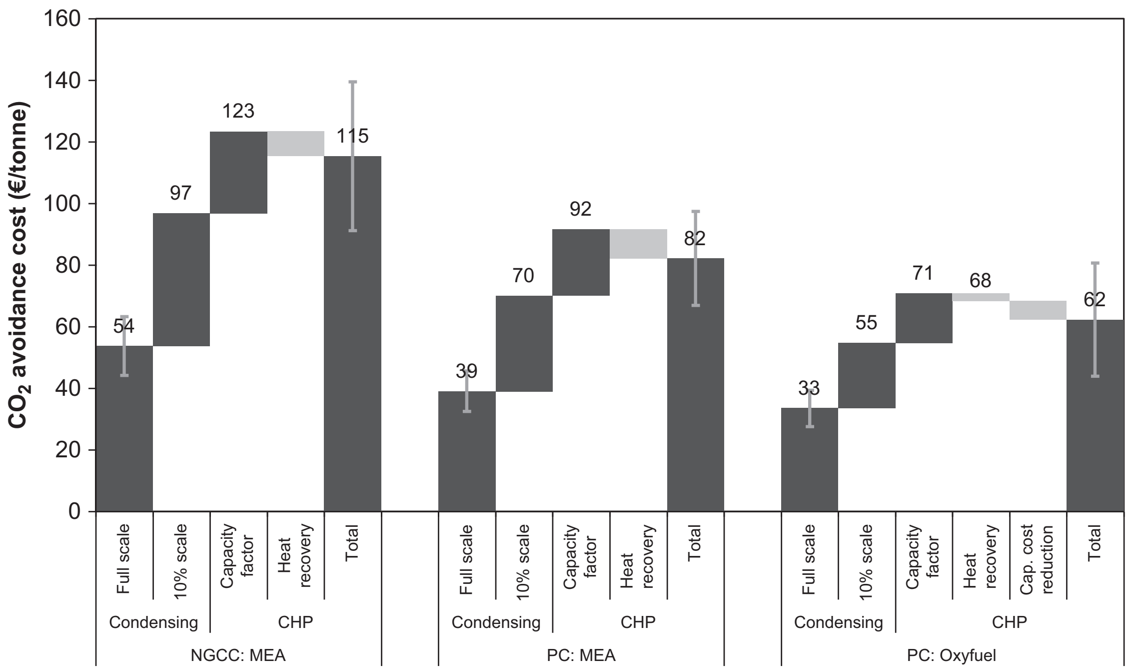


Fig. 3. Difference in CO_2 avoidance costs between full scale condensing plants (PC: 550 MW_e, NGCC: 560 MW_e) and 10% scale CHP plants caused by plant scale, capacity factor, heat recovery, and capital cost reduction. The error bars represent sensitivity of the results due to varied energy prices, interest rate, economic plant lifetime and capital costs.

to heat recovery is less than 10 €/tCO_2 avoided. The figure also shows that the lower capacity factor of CHP plants compared to centralized power plants increases the cost by more than 25 €/tCO_2 avoided.

4.3. Section summary

4.3.1. Energy penalty due to CO_2 capture in the ST

The energy penalty due to CO_2 capture ($\Delta\eta_{pen}$) was found to vary significantly by fuel type, energy system type and its application (Table 5). For natural gas-fired systems, $\Delta\eta_{pen}$ was found to range between 28% for boilers down to 11% for NGCC-CHP plants operated in low net HPR by having a good heat integration of the system. $\Delta\eta_{pen}$ was found to be larger for smaller plants because the efficiency reduction is similar for large and small scale plants in percentage point terms while the reference system efficiency of small scale plants is lower.

For coal-fired plants, $\Delta\eta_{pen}$ of 33% was observed for coal boilers without steam turbine and around 23–28% for PC power plants. The effect of scale on $\Delta\eta_{pen}$ is smaller than with natural gas-fired plants because the power plant electrical efficiency does not reduce as much as in gas-fired plants.

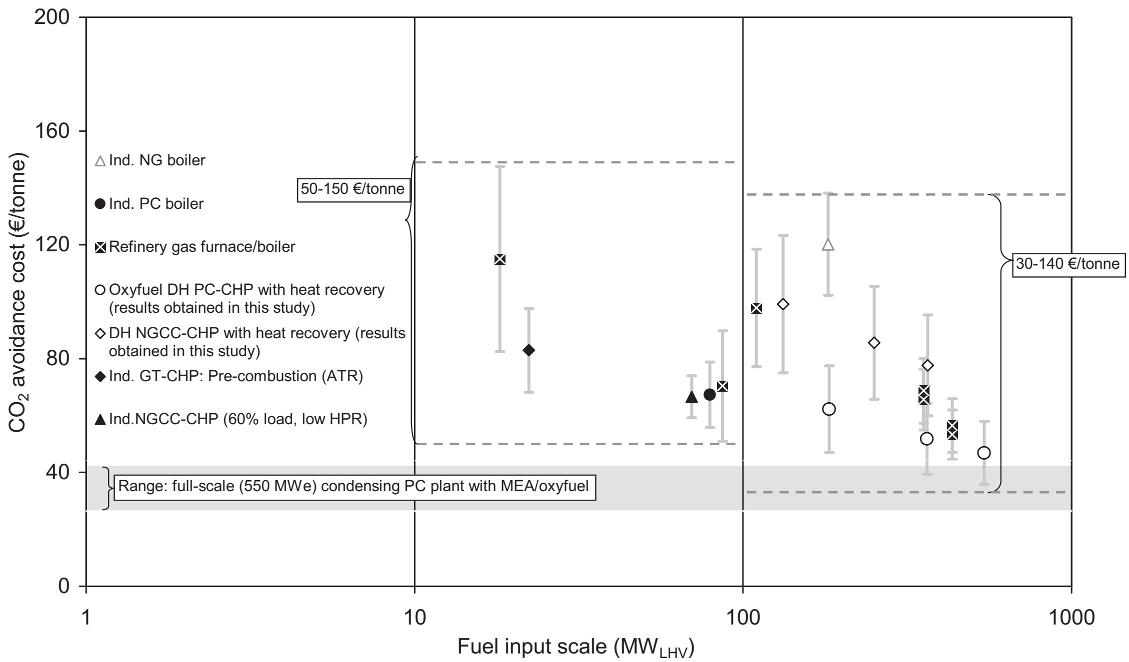


Fig. 4. CO_2 avoidance costs reported in the literature (after normalization) or calculated in this study as a function of fuel input scale (logarithmic scale) for short-term technologies.

4.3.2. CO_2 avoidance cost

Fig. 4 shows CO_2 avoidance costs reported in the literature (after normalization) and calculated in this study as a function of fuel input scale. The figure also shows the cost range for full-scale condensing power plants (29–45 €/tonne). It can be concluded that in the ST/MT, CO_2 avoidance costs range between 30 and 140 €/tonne for plant scales larger than 100 MW_{LHV} (fuel input) and 50–150 €/tonne for 10–100 MW_{LHV}. The CO_2 avoidance costs for DH–CHP plants calculated in this study are in line with the range of costs calculated for technologies presented in Table 5. From the results it is concluded that the more promising options are: (1) industrial NGCC–CHP plants with chemical absorption capture in low net HPR and low load operations, (2) oxyfuel combustion capture from PC–CHP plants, and (3) some refinery gas-fired plants with oxyfuel combustion capture.

Fig. 4 indicates a clear effect of scale on CO_2 avoidance costs, with the exception of oxyfuel combustion CO_2 capture for refinery gases. The differences observed are caused by two key reasons. First, the degree of furnace/boiler efficiency improvement by converting to oxyfuel combustion differs by furnace/boiler type. Second, the amount of oxygen required depends on the composition of the refinery gas.

The uncertainty on costs becomes significantly larger for smaller scale CO_2 capture than for large-scale capture mainly because of the higher specific capital costs. However, small scale CO_2 capture may still become competitive to large-scale capture under specific circumstances, e.g., low interest rate combined with low energy prices.

5. Assessment of long-term CO_2 capture technologies

Table 11 shows the technical and economic performance data for small–medium scale installations with CO_2 capture technologies that may become available in the LT. As noted in the introduction, this section focuses on the CO_2 capture technology itself, rather than on the combination of CO_2 capture technologies and energy system types for different applications. In many of the proposed systems, CO_2 capture process is comprised entirely or partially of membranes. As it was done in the previous section, we also assessed the applicability of large scale technologies for smaller scale emission sources.

5.1. Post-combustion capture

There are several studies in the literature that investigated the techno-economic performance of various membrane-based CO_2 capture technologies. A brief description of different types of membranes under investigation follows.

5.1.1. Ceramic membranes: Molten carbonate fuel cell as a CO_2 separator

Molten carbonate fuel cell (MCFC) may become not only a power generation technology of the future but also a cost-effective CO_2 concentrator. In an MCFC stack, CO_2 is used as cathode fuel to generate electricity and it transports to anode side of the cell in the form of carbonate ions (CO_3^{2-}). As a result, anode off-gas has a high concentration of CO_2. A large fraction of this gas is recycled back to the cathode inlet to supply CO_2 in conventional MCFC systems.[16] However, the CO_2-rich anode off-gas can also be sent to a purification and compression process if there is an external source of CO_2 that can be fed to the cathode, e.g., flue gas of combustion plants.

Several studies modeled and experimented systems to capture CO_2 from conventional combustion plant flue gas using MCFC [2,9,19,73,97,102]. A pilot plant demonstrated a stable operation of CO_2 separation from the flue gas of a coal fired power plant with a SO_2 concentration of 20–70 ppm[17] for 2000 h with little fuel cell voltage degradation, indicating a possible continuous operation of 40,000–50,000 h [102]. Although a number of concerns[18]

[16] The MCFC plant CO_2 production is ultimately contained in the cathode outlet flow where CO_2-rich gas is diluted by air [9].

[17] These values are much higher than the limit for chemical absorption capture using MEA, which is reported to be 10 ppm [15].

[18] In the literature, the following concerns have been named: (1) molten carbonate is a corrosive material operating at high temperatures, making it difficult to handle; (2) high temperatures necessary for cell operation are

Table 11
Energy, economic and CO_2 performance data for small–medium scale installations with long term CO_2 capture technologies reported in the literature.

Scale		Fuel type	Energy plant type [MW output]	Plant scale	CO_2 capture technology	CO_2 capture efficiency (%)	Conversion efficiency (η) (%)	Energy penalty ($\Delta\eta_{pen}$)	Capital cost [€/kW output] (%)	O&M cost [%-capital cost]	Reference
Condensing/CHP plants	Medium (5–50 MW_e)	Natural gas	SOFC-CHP (industrial)	5	Oxyfuel (stand-alone ASU)	94 [a]	53	5	2100 [b]	6	[69]
			SOFC-GT-ST	20	Oxyfuel (integrated OCM)	100	59	–[c]	2400	3	[71]
			AZEP (condensing)	50	Oxyfuel (membrane-based)	85–100	49–50	5–9	1900–2100	4	[74]
	Small (0.005–<5 MW_e)	Natural gas	SOFC-CHP (DH)	0.5	OCM	100	52	5	2500	4	[55]
			SOFC-GT		Oxyfuel (stand-alone ASU)	100	58	4 [d]	1600	4	[32]
Boilers	Medium (5–50 MW_e)	Natural gas	Boiler (industrial and DH)	170	Oxyfuel (OCM)	92	93	4	130–360	5	[100,55]
		Coal	Boiler (industrial)	50	Oxyfuel (OCM)	86–98	81	6–8	560–680	4	[55]
H_2 plants	Small (0.005–5 MW_e)	Natural gas	WGSMR (refueling, DH)	1.7	Pre-combustion	57	69	4	770 [b]	15	[95]
		Natural gas	WGSMR (refueling, DH): current technology level	0.9	Pre-combustion	50	67	9	2800	N.A.	[70]
Others	Various	Various	MCFC using external CO_2 as cathode fuel	4	MCFC (electro-chemical)	100–200	26–31	21–36	2700–2900	16	Own estimate based on [31] [e]

[a] The reported CO_2 purity is 84%. Therefore, the CO_2 capture efficiency was recalculated as described in Section 3.4.
[b] Compressor costs are recalculated using Eq. (9).
[c] The reference literature compares the technical performance with a conventional NGCC of identical scale with an electrical conversion efficiency of 40% [71]. We considered the comparison to be unfair because the state-of-the-art NGCC of similar scale (L20A-1, 25 MW_e net) already shows 48.6% efficiency (LHV) [37].
[d] The referenced study does not discuss the lifetime of fuel cell stacks and the costs for cell stack replacement. Therefore, it is not clear how these costs are taken into account. The referenced study also does not provide separate cost figures for variable O&M costs (chemicals, water, etc.). These costs, however, are reported to be of limited influence on total costs [89].
[e] Frangini [31] estimates that the electrical efficiency of MCFC decreases by 10% when using coal combustion flue gas and by 30% when using natural gas combustion flue gas compared to the conventional operation of recycling the anode off-gas as cathode fuel. While the Frangini study uses the current cost for the MCFC system, we assumed that the capital cost decreases to one-tenth of the current level in the LT. This is equivalent to around 1200 €/kW_e without CO_2 capture, which is roughly in line with the predictions presented in Araki and Keppo [4]. The electrical efficiency is assumed unchanged in the LT.

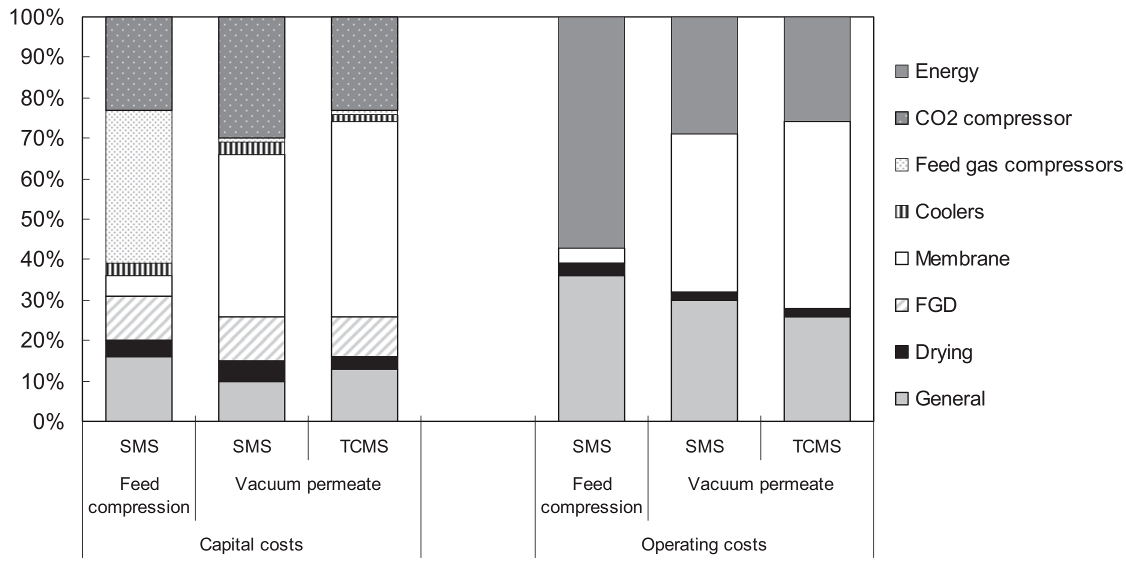


Fig. 5. Breakdown of capital costs and operating costs for CO_2 separating polymeric membranes operating under vacuum permeate and feed compression conditions. SMS: single stage membrane system, TCMS: two stage cascade membrane system. *Source*: Adapted from Ho et al. [41].

over the practicality of CO_2 capture from flue gas using MCFC has been expressed [87], this unique combination of power generation and "active" CO_2 concentration (as opposed to "passive" CO_2 concentration, which consumes large amount of energy) could become a viable solution in the future [11].

5.1.2. Polymer-based membranes

CO_2 capture using commercially available polymeric membranes have resulted in higher energy penalties compared to chemical absorption processes [59], but significant performance improvement is expected in the future. Polymeric membrane CO_2 separation is driven by the partial CO_2 pressure difference between the feed and permeate gases by either using a compressor on the feed side or a vacuum-pump on the permeate side.

Polymeric membrane separation may become economically attractive for smaller scale emission sources although there is no feasibility study for smaller scale applications available. This is partly because many studies on polymeric membrane-based CO_2 separation focus on coal combustion due to the high CO_2 concentration in the flue gas. Literature suggests that the use of vacuum pumps to increase the partial pressure difference reduces energy consumption considerably[19] [28,41,114] and possibly total capital costs [41]. The fraction of membranes in the total capital costs (39–47%) is considerably larger (5%) for vacuum pump-driven systems than for compressor-driven systems (see Fig. 5). This is favorable for smaller scale CO_2 capture because the capital cost increase resulting from the scale-down of the plant may be limited.

5.1.3. Facilitated transport membranes

Matsumiya et al. [76] report that facilitated transport membranes show better energy and economic performance than polymer membranes. The essential element of facilitated transport membranes is the specific chemical interaction of a gas component with a compound present in the membrane, the so-called carrier [59]. Carbonates, amines and molten salt hydrates have been suggested as carriers for CO_2 separation. The separation process is driven by the difference in partial pressure of the component to be transported [59].

5.1.4. Membrane-gas hybrid processes

Membrane gas absorption (MGA) combines the advantages of absorption (high selectivity) and membrane technologies (modularity and small size) [29]. A porous, water repellent, polymeric membrane separates the gas phase (feed gas) from the liquid phase (absorbent). CO_2 in the flue gas diffuses through the membrane and gets absorbed into the liquid absorbent. The CO_2-rich absorbent is regenerated by heating it in a stripper, just as in a conventional chemical absorption process. It is suggested that the MGA process may enable lower CO_2 capture cost compared to conventional chemical absorption process at smaller plant scales due to its modular nature [55]. There are, however, studies that suggest otherwise; capital costs are dominated by components other than the membrane module such as the CO_2 stripper tower [35,111].[20]

The membrane flash process investigated by RITE e.g., [76,85] may become suitable for small-scale applications. In this process, the feed gas and carrier solution are both supplied to the high-pressure feed side of the hollow fiber membrane module. Driven by the vacuum pump at the permeate side, the CO_2-rich solution permeates the membrane and liberates the dissolved CO_2 at the permeate side of the membrane. The system proposed by Matsumiya et al. [76] uses neither absorber nor desorber towers. Therefore, the membrane module is reported to account for about 60% of total capital costs (excluding FGD, including CO_2 purification and liquefaction) [76].

(*footnote continued*)
extremely difficult for integration into a post-combustion application; (3) only small voltage is allowable to avoid decomposition of the molten carbonate, leading to a huge stack requirement; (4) sensitivity to the contaminants such as SO_2.

[19] The required membrane area becomes much larger for vacuum pump-driven systems due to the lower pressure ratio, but the consequent cost increase could be offset by the omission of expensive feed gas compressor.

[20] Grønvold et al. [35] shows that the fraction of membrane contactor in total capital cost (excluding FGD and including CO_2 compression unit) is only 14%. The cost estimates from Yan et al. [111] shows that also for a smaller plant scale of 2.9 kg CO_2/s captured, the fraction of membrane contactor in total capital cost (excluding both FGD and CO_2 compression units) is less than 30%.

5.2. Oxyfuel combustion capture

Oxygen conducting membranes (OCM) is a promising advanced air separation technology. OCM is a ceramic membrane made of perovskites, which have both electronic and oxygen ionic conductivity when operated at high temperatures around 700 °C to 900 °C [3]. Oxygen ions are transferred under a gradient of oxygen partial pressure on the opposite side of the membrane. OCM can produce oxygen of very high purity, above 99% [96]. A high level of system integration can be achieved for gas-fired boilers. In a gas-fired boiler system proposed by Switzer et al. [100], the gradient of oxygen partial pressure is generated by placing a fuel combustor at the permeate side of the membrane.

Another oxyfuel combustion-based technology that may become economically competitive for smaller scale gas-fired power generation is the advanced zero emission power (AZEP) concept. A general description of the AZEP technology can be found in, e.g., [21,74,98]. The key component is a mixed conducting membrane (MCM) reactor. Compressed air flows through the MCM reactor, where oxygen is separated and directed to the combustion chamber. The combustion heat is transferred to O_2-depleted compressed air via the MCM reactor. Electricity is generated from an expansion of the hot O_2-depleted compressed air in a gas turbine and an expansion of steam generated from the gas turbine exhaust gas, just like in a conventional NGCC plant. A modeling study indicates that the AZEP plant may perform better at smaller scales (about 50 MW_e) than at large scales (around 400 MW_e) because of the smaller difference in combustor outlet temperature between the reference NGCC and AZEP plant resulting in smaller energy penalty [74]. Although there are still technical challenges in the MCM development [21], it may become an attractive option for medium-size DG.

5.3. Pre-combustion CO_2 capture: Hydrogen plants

Small-scale H_2 production plants may serve as refueling stations for future hydrogen vehicles as well as for district and residential fuel cell systems. Water gas shift membrane reactor (WGSMR) technology suits well for these applications because of its modular nature. In addition, WGSMR technology is expected to enable cheap CO_2 capture [95]. Tokyo Gas has been running 120 kW scale demonstration units since 2004 [70]. CO_2 avoidance cost estimates for 2020 are reported to be well above 100 €/tonne [70], but the long-term durability of the membranes still remains a challenge [93]. We therefore did not consider the cost figures reported by [70] to be representative of commercial small-scale H_2 generation technologies either in the ST/MT or the LT. Nevertheless, the cost estimates reported in Sjardin et al. [95] shown in Table 11 should be achievable in the longer term.

5.4. Fuel cell systems: Solid oxide fuel cells (SOFC)

Solid oxide fuel cell (SOFC) consists of electrolytes made of a solid, non-porous metal oxide (Y_2O_3-stabilized ZrO_2), and it operates at 600–1000 °C where ionic conduction by oxygen ions take place [25]. SOFC is seen as a promising future energy conversion technology for DG systems [107] for two reasons. First, SOFC enables high electrical conversion efficiency for both large and small scale systems. In particular, SOFC hybrid systems, which integrate SOFC with GT or steam turbine, may achieve electrical conversion efficiency as high as 70% (an overview of literature can be found in, e.g., [107]). Second, SOFC has a potential to enable cheap CO_2 capture. In a SOFC system, oxygen is transported through the electrolyte from the cathode side to the anode side, where the fuel is oxidized. Therefore, for natural gas-fed SOFC systems, CO_2 contained in the anode off-gas as a result of reforming is not diluted by nitrogen and is potentially easy to capture.

An overview of capture technologies applicable for SOFC systems can be found in, e.g., Kuramochi et al. [69][21] and Damen et al. [21]. A review study by Wee [107] suggests that the SOFC hybrid systems of 1–100 MW_e scale may achieve electrical conversion efficiency higher than 60% even when equipped with CO_2 capture. Moreover, recent literature [18] indicates that an11 MW_e coal-powered CO_2 capture SOFC system comprised of chemical looping hydrogen generation process and SOFC/gas turbine cycle can achieve a high net electrical efficiency of more than 43%.

Although SOFC is currently considered to be too expensive for commercialization, the potential for cheap CO_2 capture from advanced fuel cell technologies such as SOFC may facilitate the market introduction of these fuel cells in a carbon-constrained society [68]. To date, however, there are only a few studies that explored the techno-economic performance of CO_2 capture from SOFC systems ([32,55,69,71], see Table 11).

5.5. Results

As seen in Table 11, the energy penalty for advanced small–medium scale energy conversion technologies due to CO_2 capture compared with the identical plant without CO_2 capture ($\Delta\eta_{pen}$) reported in the literature ranged between 4% and 9% for CO_2 capture efficiency of 85–100%. These values are significantly smaller than those reported for ST/MT technologies because the advanced systems are designed to capture CO_2.

Fig. 6 shows CO_2 avoidance costs for distributed energy systems based on advanced technologies as a function of fuel input scale. Note that the performance of SOFC–GT–ST hybrid system [71] was compared with that of the reference NGCC plant (50 MW_e) used for the AZEP power technology because the performance of an identical SOFC–GT–ST plant without CO_2 capture is not reported in the referenced study [71]. The calculations were performed for different capacity factors: 7500 h/yr and 5000 h/yr. Note that these cost estimates assume that the technologies are mature. CO_2 avoidance costs range around 10–90 €/tonne for plant scales larger than 100 MW_{LHV}, 25–100 €/tonne for 10–100 MW_{LHV} and 35–150 €/tonne for 10 MW_{LHV} or smaller.

The impact of capacity factor assumption on cost performance is significant (10–30 €/tonne increase in CO_2 avoidance cost) for some capital-intensive technologies that competes economically with conventional technologies by demonstrating a considerably higher energy efficiency, e.g., SOFC systems and H_2 membrane reactors. For boilers, the impact is limited because of limited additional capital costs. The economies of scale is not clear from Fig. 6; there are a number of possible explanations for this result. First, some technologies reported for smaller scale systems may be more advanced than those reported for larger scale systems. Second, there is more uncertainty in the performance data of LT technologies because they are studied in less detail compared to ST/MT CO_2 capture technologies, for which case-specific feasibility studies are available.

6. Combined results

Fig. 7 combines the short- and long-term CO_2 avoidance cost results presented in Figs. 4 and 6. The two colored bands roughly show the range of CO_2 avoidance costs for short-term and long-term technologies, respectively. In the ST/MT, CO_2 avoidance costs would range between 40 and 160 €/tonne for plant scales larger than 100 MW_{LHV}, 60–200 €/tonne for 10–100 MW_{LHV} and 80–200 €/tonne

[21] The referenced study defines the timeframe "up to 2025". However, taking into account the recent developments for commercialization of SOFC, it is more likely that CO_2-capture SOFC systems will be introduced to the market after 2030.

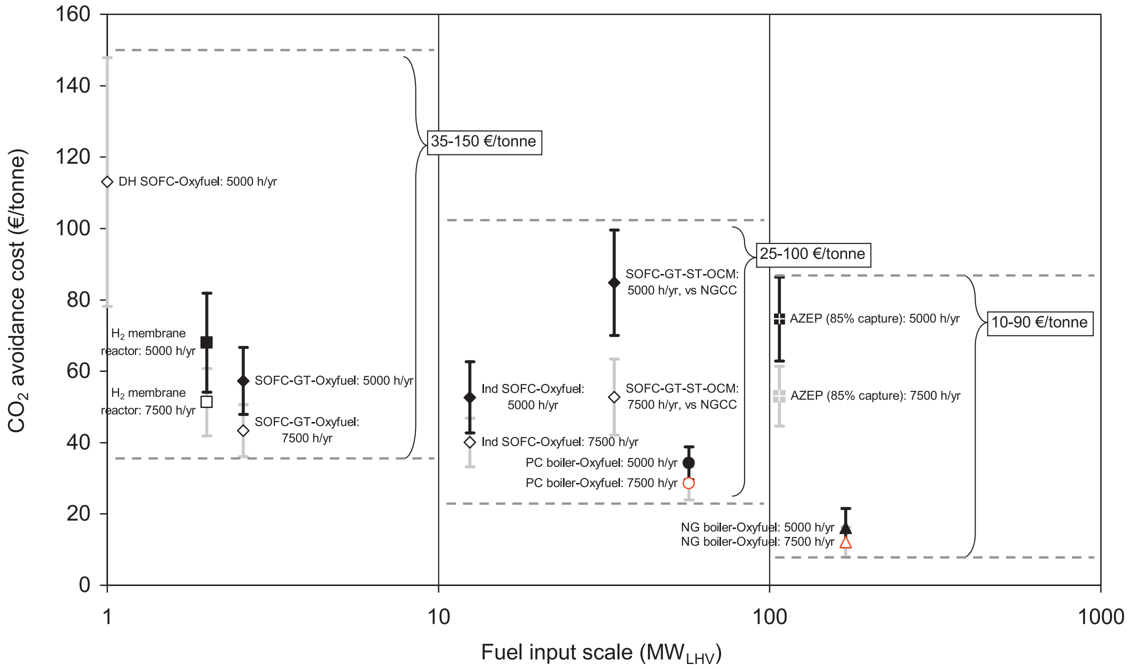


Fig. 6. CO_2 avoidance costs reported in the literature as a function of fuel input scale (logarithmic scale) for long-term technologies. It is shown that the reduced load leads to considerably higher costs.

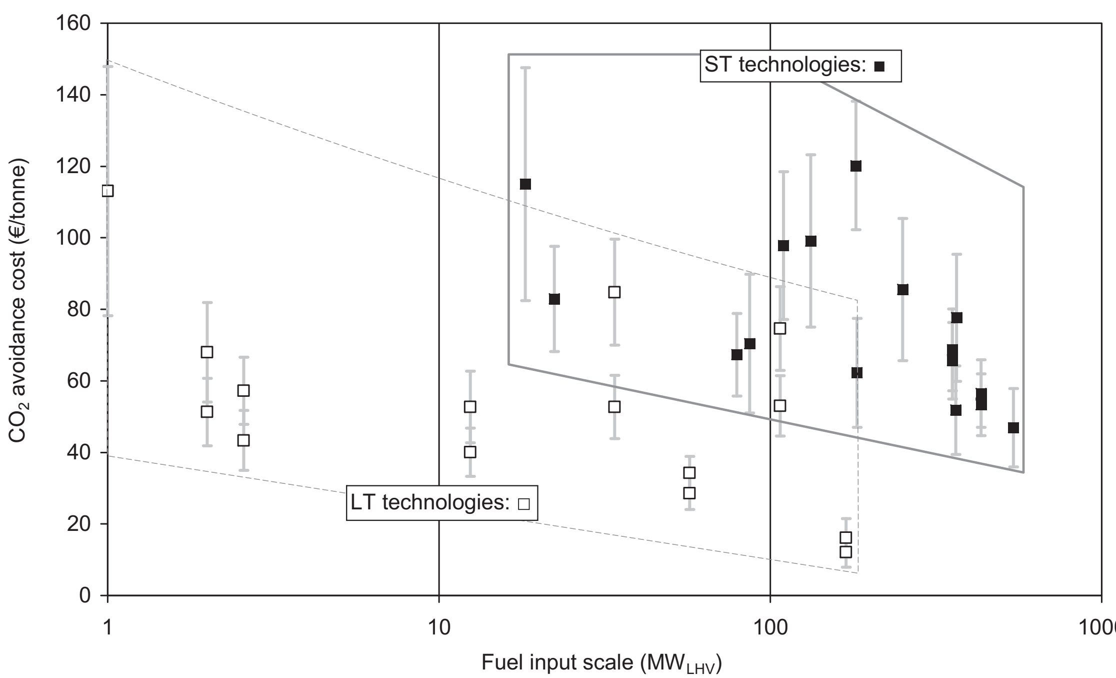


Fig. 7. CO_2 avoidance costs reported as a function of fuel input scale (logarithmic scale) for short-term and long-term technologies. The clusters and its borders are only for illustrative purposes.

or more for 10 MW_{LHV} or smaller. For the LT, it was estimated that the CO_2 avoidance costs may range between 10 and 40 €/tonne for plant scales larger than 100 MW_{LHV}, 10–80 €/tonne for 10–100 MW_{LHV} and 10–120 €/tonne for 10 MW_{LHV} or smaller.

7. CO_2 compression and distributed transport costs

In this section, the costs of CO_2 compression and distributed transport between the CO_2 capture site and trunk pipeline using branch pipeline and truck are evaluated. The schemes of two transport options compared were presented in Fig. 2 and the parameters used for the calculations were presented in Table 4. Note that the assumptions on the transport distance (30 km) and the lack of network for distributed CO_2 collection is conservative. Many distributed emission sources would be located within a distance far less than 30 km from the nearest trunk pipeline once a large-scale CO_2 transport infrastructure is developed.

Fig. 8 shows the CO_2 compression and distributed transport costs calculated for the two options while Fig. 9 shows the

breakdown of the costs by component for different transport scales. Truck transport became cheaper than pipeline transport for emission sources smaller than around 200 tCO_2/d (equivalent to combustion plants of 20–30 MW_{LHV} natural gas input) because of lower investment costs. As shown in Fig. 9, the costs were dominated by CO_2 compression (65% at the lowest) and the cost for transport itself is small (35% at the highest) for all scales. For larger scales when branch pipeline transport is the cheapest, the branch pipeline cost was about 5 €/tonne (see 500 t/d in Fig. 9). For smaller scales, when truck transport is the cheapest, the cost was about 6 €/tonne and is unaffected by the scale (see 5 and 100 t/d in Fig. 9). Fig. 9 also shows that the CO_2 recompression cost for the truck transport option is negligible.

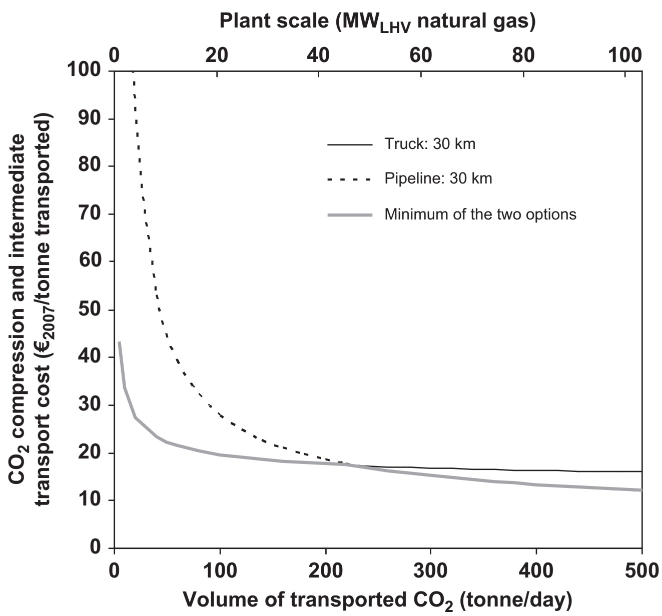


Fig. 8. Small scale CO_2 transport costs as a function of transported CO_2 volume per day.

Our results were in line with those from IEA GHG study [56], which reports a marginal cost of 34 $/$tCO_2$ transported (about 25 $€_{2007}/tCO_2$) for collecting 0.7 $MtCO_2$/yr from 49 sources (average 14 $ktCO_2$/yr per source). This referenced study assumed a branch transport network using high density polyethylene (HDPE) pipes with the CO_2 transport driven by vacuum pumps.

8. Discussion

The results obtained in this study and their main uncertainties and sensitivities have been discussed in their respective sections. In this section, we mainly discuss the limitations of this study and the implications of the obtained results in relation to the expected future CO_2 price.

A number of limitations of this study are identified. First, the literature survey showed that there are only a few studies available on the techno-economic feasibility of CO_2 capture from distributed energy systems. For many energy conversion technologies with CO_2 capture, there was only one techno-economic study available. Therefore, it was difficult to assess whether the reported values are reliable. Moreover, some of the reported economic performance data may be outdated. Furthermore, many studies do not have clear definition of "capital cost" because these studies are of an order-or-magnitude analysis nature.

Second, the results obtained in this study may be sensitive to the reference technologies selected for comparison. In this study, we compared identical energy conversion plants with and without CO_2 capture. However, since DG systems will always be compared with centralized generation systems, the economic competitiveness of CO_2 capture-fitted DG systems will likely depend largely on the local energy market circumstances of the CO_2 capture site. Moreover, advanced energy conversion systems investigated in this study are designed to capture CO_2 and it may not make economical sense to apply the technology without CO_2 capture. In such a case, the competitor of an advanced energy

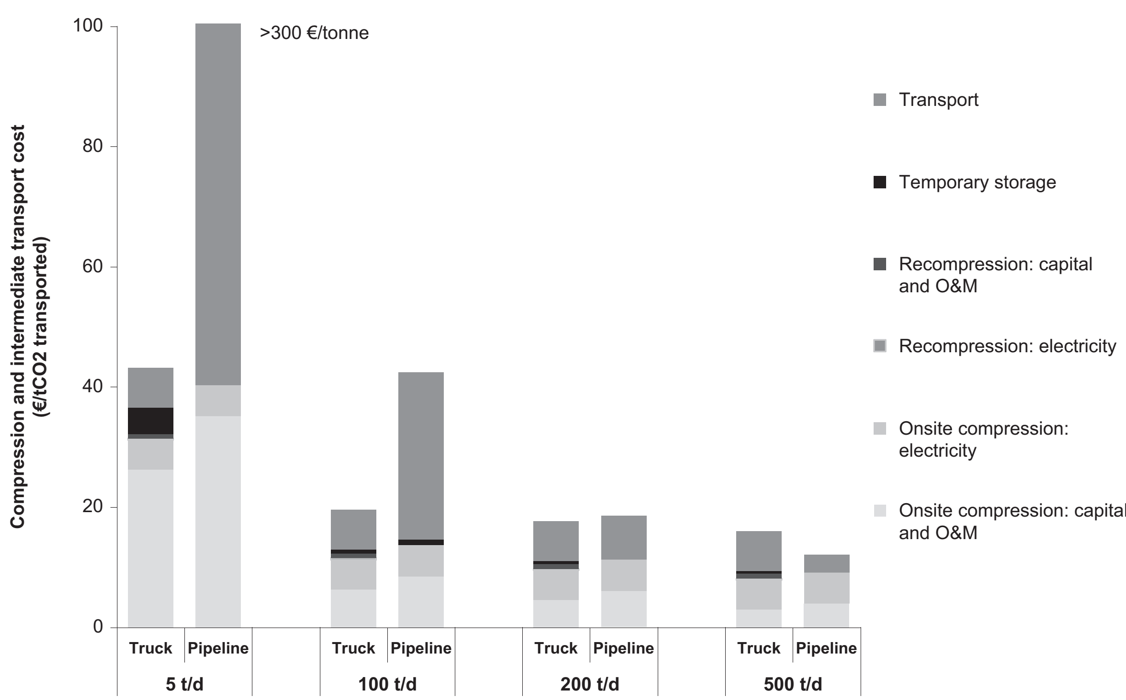


Fig. 9. Breakdown of CO_2 compression and distributed transport costs.

system with CO_2 capture would be conventional energy conversion systems and therefore, the CO_2 mitigation performance results may differ significantly from those reported in this study.

The third limitation relates to the nature of distributed energy systems. In this study, the uncertainty of CO_2 avoidance costs was found to be considerably larger than that for large-scale centralized power plants. This is logical because decentralized energy conversion systems show a much larger diversity in their operational conditions (e.g., load factor and heat-to-power ratio) compared to centralized power plants.

9. Conclusions

CO_2 emissions from distributed energy systems may become significant; accounting in 2030 for about 20% of the current global energy-related CO_2 emissions. This study assessed and compared the technical and economic performance of CO_2 capture technologies for distributed energy systems. We took into account the diversity in type of generator technologies used, applications, plant scales, operational patterns and timeframe in which the technologies may become commercial. We also investigated the compression and distributed transport of captured CO_2 between the capture site and trunk pipeline. The study covered CHP plants, boilers and distributed hydrogen plants. The following conclusions are drawn:

In the ST/MT, the energy penalty due to CO_2 capture ($\Delta\eta_{pen}$) varies significantly by fuel type, energy system type and its application. For natural gas-fired installations, $\Delta\eta_{pen}$ was found to range between 28% for boilers down to 11% for NGCC-CHP plants. For coal-fired installations, $\Delta\eta_{pen}$ of 33% was found for coal boilers and around 23–28% for PC power plants. It can be concluded that in the short-term (around 2020–2025), CO_2 avoidance costs range between 30 and 140 €/tonne for plant scales larger than 100 MW_{LHV} (fuel input) and 50–150 €/tonne for 10–100 MW_{LHV}. The most promising smaller-scale options identified are: (1) industrial NGCC–CHP plants with chemical absorption capture in low net HPR and low load operations, (2) oxyfuel combustion capture from PC–CHP plants, and (3) oxyfuel combustion capture from some refinery gas-fired plants.

We also investigated the heat integration potentials between a power plant and a CO_2 capture process. For chemical absorption capture, it was suggested that up to 30% of the solvent regeneration heat can be recovered in the form of hot water for DH without additional capital cost. For oxyfuel combustion capture from coal-fired plants, it was suggested that 30% of the energy consumption for ASU and CO_2 compression can be recovered and about 5% reduction of capital cost can be achieved by heat integration. The resulting reduction of CO_2 avoidance costs, however, was found to be up to 10 €/tCO_2 for post-combustion capture and around 3 €/tCO_2 for coal-fired oxyfuel capture.

In the long-term future, a number of advanced CO_2 capture technologies using oxygen separation membranes have been found for both coal and natural gas-fired systems. Membrane-based oxyfuel combustion processes increases the furnace/boiler efficiency without significant electricity consumption. Other promising technologies include solid oxide fuel cells (SOFC) for power and CHP applications, and water–gas shift membrane reactor (WGSMR) for hydrogen production. For coal-fired power plants, there are also a number of advanced post-combustion capture technologies that may enable low-cost CO_2 capture at small scales, including modified molten carbonate fuel cells (MCFC) and polymer-based membranes.

The energy penalty due to CO_2 capture ($\Delta\eta_{pen}$) was found to range around 4–9%. Regarding the economic performance, it was concluded that in the LT, CO_2 avoidance costs would be around 10–90 €/tonne for plant scales larger than 100 MW_{LHV}, 25–100 €/tonne for 10–100 MW_{LHV} and 35–150 €/tonne for 10 MW_{LHV} or smaller. The influence of the capacity factor is significant for some technologies that demonstrate considerably higher energy efficiency than the conventional technologies but are also more capital cost-intensive. The economies of scale on CO_2 avoidance costs was not clear due to lack of data points and the possibility that some smaller scale technologies are more advanced than the larger scale technologies reported.

Our results also show that CO_2 compression and distributed transport costs could be significant. For the distance of 30 km, 10 €/tonne was calculated for scales below 500 tCO_2/day and more than 50 €/tonne for scales below 5 tCO_2/day (equivalent to 1 MW_{LHV} natural gas). CO_2 compression accounted for the vast majority of these costs. Transport distance was found to have limited influence. The results indicate that the reduction of CO_2 compression cost, which is closely linked to the CO_2 transport mode, may become a key to economical CO_2 capture for smaller scale energy systems.

The results of this article highlight the possibilities for cost-effective CO_2 capture in distributed energy systems in case CO_2 transport and storage infrastructure has been developed. The costs of CO_2 capture are not prohibitively expensive even in the ST/MT and have a significant cost reduction potential in the LT. CCS from distributed emission sources would also contribute to the economies of scale of CO_2 transport and storage. However, further research is required to have better understanding on the potential of CO_2 capture in distributed energy systems. Examples are: (1) analysis on the minimization of CO_2 compression and distributed transport costs by comparing various system configurations, and (2) exploring the possibilities to extend annual operation time for DH CHP plants to enable more cost-effective CO_2 capture.

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